

Sertifikaat

PATENTKANTOR REPUBLIC OF SOUTH AFRICA



IB/03/04733

Certificate

REPUBLIEK VAN SUID-AFRIKA PATENT OFFICE

DEPARTEMENT VAN HANDEL
EN NYWERHEID

DEPARTMENT OF TRADE AND
INDUSTRY

REC'D 04 MAR 2004

WIPO PCT

Hiermee word gesertifiseer dat
This is to certify that

the documents annexed hereto are true copies of:

Application form P.1 and P.3, provisional specification and
drawings of South African Patent Application No. 2002/8708 ✓
as originally filed in the Republic of South Africa on 28 October ✓
2002 in the name of **SASOL TECHNOLOGY (PROPRIETARY)**
LIMITED for an invention entitled: **"PRODUCTION OF SYNTHESIS**
GAS AND SYNTHESIS GAS DERIVED PRODUCTS".

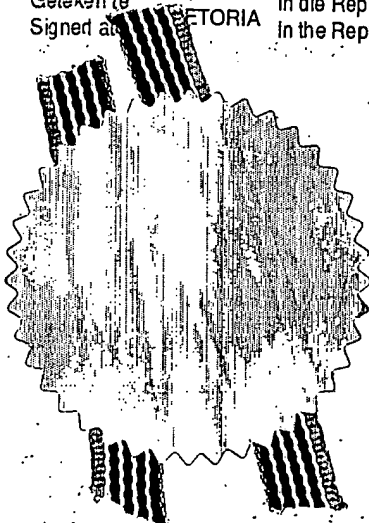
PRIORITY DOCUMENT
SUBMITTED OR TRANSMITTED IN
COMPLIANCE WITH
RULE 17.1(a) OR (b)

Geteken te
Signed at VICTORIA in die Republiek van Suid-Afrika, hierdie
in the Republic of South Africa, this

10th

dag van
day of December 2003

Registrateur van Patente



REPUBLIC OF SOUTH AFRICA
PATENTS ACT, 1978
APPLICATION FOR A PATENT AND
ACKNOWLEDGEMENT OF RECEIPT
(Section 30(1) Regulation 22)

REPUBLIC OF SOUTH AFRICA	
FORM P.1	REVENUE
(to be lodged in duplicate)	
28.10.02 ✓	R 060.00
REPUBLIC VAN SUID AFRIKA	
A&A REF	15463 56

THE GRANT OF A PATENT IS HEREBY REQUESTED BY THE UNDERMENTIONED APPLICANT
ON THE BASIS OF THE PRESENT APPLICATION FILED IN DUPLICATE

21 01 PATENT APPLICATION NO 2002/8708

71 FULL NAME(S) OF APPLICANT(S)

SASOL TECHNOLOGY (PROPRIETARY) LIMITED

ADDRESS(ES) OF APPLICANT(S)

1 Sturdee Avenue, Rosebank, Johannesburg, Gauteng Province, Republic of South Africa

54 TITLE OF INVENTION

PRODUCTION OF SYNTHESIS GAS AND SYNTHESIS GAS DERIVED PRODUCTS

Only the items marked with an "X" in the blocks below are applicable.

☐ THE APPLICANT CLAIMS PRIORITY AS SET OUT ON THE ACCOMPANYING FORM P.2. The earliest priority claimed is

Country:

No:

Date:

☐ THE APPLICATION IS FOR A PATENT OF ADDITION TO PATENT APPLICATION NO 21 01

☐ THIS APPLICATION IS A FRESH APPLICATION IN TERMS OF SECTION 37 AND BASED ON
APPLICATION NO 21 01

THIS APPLICATION IS ACCOMPANIED BY:

- | | |
|---|--|
| X | A single copy of a provisional specification of 16 pages |
| X | Drawings of 1 sheet |
| | Publication particulars and abstract (Form P.8 in duplicate) (for complete only) |
| | A copy of Figure of the drawings (if any) for the abstract (for complete only) |
| | An assignment of invention |
| | Certified priority document(s). (State quantity) |
| | Translation of the priority document(s) |
| | An assignment of priority rights |
| | A copy of Form P.2 and the specification of RSA Patent Application No 21 01 |
| X | Form P.2 in duplicate |
| | A declaration and power of attorney on Form P.3 |
| | Request for ante-dating on Form P.4 |
| | Request for classification on Form P.9 |
| | Request for delay of acceptance on Form P.4 |
| | Extra copy of informal drawings (for complete only) |

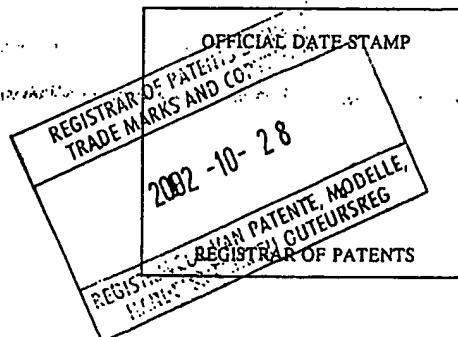
74 ADDRESS FOR SERVICE: Adams & Adams, Pretoria

Dated this 28TH day of OCTOBER 2002

ADAMS & ADAMS
APPLICANTS PATENT ATTORNEYS

The duplicate will be returned to the applicant's address for service as
proof of lodging but is not valid unless endorsed with official stamp

A&A P201



ADAMS & ADAMS
PRETORIA

REPUBLIC OF SOUTH AFRICA
PATENTS ACT, 1978
DECLARATION AND POWER OF ATTORNEY
(Section 30 - Regulation 8, 22(i)(c) and 33)

FORM P.3

PATENT APPLICATION NO		
21	01	2002/8708

A&A Ref: V15463

LODGING DATE	
22	28 OCTOBER 2002

FULL NAME(S) OF APPLICANT(S)	
71	SASOL TECHNOLOGY (PROPRIETARY) LIMITED

FULL NAME(S) OF INVENTOR(S)	
72	KEYSER, Martin Jakobus COERTZEN, Margaretha

EARLIEST PRIORITY CLAIMED	COUNTRY	NUMBER	DATE			
	33	NIL	31	NIL	32	NIL

NOTE: The country must be indicated by its International Abbreviation - see schedule 4 of the Regulations

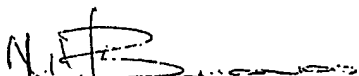
TITLE OF INVENTION	
54	PRODUCTION OF SYNTHESIS GAS AND SYNTHESIS GAS DERIVED PRODUCTS

I/we **BARRADAS, Morné**

hereby declare that :-

1. ~~I/we am/are the applicant(s) mentioned above;~~
- ** 2. I/we have been authorized by the applicant(s) to make this declaration and have knowledge of the facts herein stated in the capacity of **AUTHORISED SIGNATORY** of the applicant(s);
- *** 3. the inventor(s) of the abovementioned invention is/are the person(s) named above and the applicant(s) has/have acquired the right to apply by virtue of an assignment from the inventor(s);
4. to the best of my/our knowledge and belief, if a patent is granted on the application, there will be no lawful ground for the revocation of the patent;
- **** 5. ~~this is a convention application and the earliest application from which priority is claimed as set out above is the first application in a convention country in respect of the invention claimed in any of the claims; and~~
6. the partners and qualified staff of the firm of ADAMS & ADAMS, patent attorneys, are authorised, jointly and severally, with powers of substitution and revocation, to represent the applicant(s) in this application and to be the address for service of the applicant(s) while the application is pending and after a patent has been granted on the application.

SIGNED THIS 29 DAY OF November 2002



Company Name: **SASOL TECHNOLOGY (PROPRIETARY) LIMITED**

Full Names: **BARRADAS, Morné**

Capacity: **AUTHORISED SIGNATORY**

(no legalization necessary)

- * In the case of application in the name of a company, partnership or firm, give full names of signatory/signatories, delete paragraph 1, and enter capacity of each signatory in paragraph 2.
- ** If the applicant is a natural person, delete paragraph 2.
- *** If the right to apply is not by virtue of an assignment from the inventor(s), delete "an assignment from the inventor(s)" and give details of acquisition of right.
- **** For non-convention applications, delete paragraph 5.

A & A Ref No: V15463

FORM P6

ADAMS & ADAMS
PATENT ATTORNEYS
PRETORIA

REPUBLIC OF SOUTH AFRICA
Patents Act, 1978

PROVISIONAL SPECIFICATION

(Section 30 (1) - Regulation 27)

21	01	OFFICIAL APPLICATION NO
----	----	-------------------------

22	LODGING DATE
----	--------------

2002/8708

28 OCTOBER 2002

71	FULL NAME(S) OF APPLICANT(S)
----	------------------------------

SASOL TECHNOLOGY (PROPRIETARY) LIMITED

72	FULL NAME(S) OF INVENTOR(S)
----	-----------------------------

KEYSER, Martin Jakobus
COERTZEN, Margaretha

54	TITLE OF INVENTION
----	--------------------

PRODUCTION OF SYNTHESIS GAS AND SYNTHESIS GAS DERIVED PRODUCTS

THIS INVENTION relates to the production of synthesis gas and synthesis gas derived products. In particular, it relates to a process for producing synthesis gas and to a process for producing a synthesis gas derived product.

Gasifiers are used to gasify solid carbonaceous materials in order to generate synthesis gas. When some gasifiers, such as Lurgi (trade name) moving bed gasifiers are used to generate synthesis gas from coal, a significant amount of methane is produced and tar and solids are present in the raw synthesis gas produced by the gasifiers. In many processes, such as a Fischer-Tropsch synthesis gas conversion process, relying on a synthesis gas feedstock produced in gasifiers from solid carbonaceous material, the methane in the synthesis gas poses a problem, as it passes through the process inertly, effectively reducing equipment capacity. The tar and solids also pose a problem as, in order to get rid of them, the synthesis gas has first to be quenched. The effect of this is that the steam generation capacity of the synthesis gas is reduced as a result of the thermal energy lost during quenching, allowing low pressure steam only to be generated.

According to one aspect of the invention, there is provided a process for producing synthesis gas, the process including

in a gasification stage, gasifying a carbonaceous feedstock to provide a raw synthesis gas which includes at least H_2 , CO and CH_4 ; and

in a partial oxidation stage, subjecting the raw synthesis gas to partial oxidation in the presence of oxygen to provide an upgraded synthesis gas which includes less CH_4 and more H_2 and CO than the raw synthesis gas.

The partial oxidation stage is typically a non-catalytic, thermal partial oxidation stage. Typically, when the carbonaceous feedstock is coal or the like, the raw synthesis gas includes tar and solids. Advantageously, by subjecting the raw synthesis gas to thermal partial oxidation, the upgraded synthesis gas which is produced is substantially free of tars and solids, obviating the need to quench the raw synthesis gas with water to get rid of the tars and solids.

The thermal partial oxidation may be effected at a temperature of between about 1000 °C and about 1600 °C, more preferably between about 1100 °C and about 1400 °C, e.g. about 1300 °C.

The process may include subjecting the upgraded synthesis gas to a water-gas shift reaction stage to provide a hydrogen enriched synthesis gas with a more desirable ratio of H₂ and CO, e.g. a ratio of 1 : 1.7 or 1 : 2. Subjecting the upgraded synthesis gas to a water-gas shift reaction stage typically includes adding steam at a temperature of at least 400 °C, preferably between about 410 °C and about 450 °C, to the upgraded synthesis gas.

Typically, the carbonaceous material is gasified in the presence of oxygen and steam in a gasifier, such as a Lurgi (trade name) moving bed gasifier.

The process may include, in a cooling stage, cooling the hydrogen enriched synthesis gas and producing steam. Steam at a pressure of at least 34 bar, preferably at least 41 bar, is produced in the cooling stage. This high pressure steam may advantageously be used in the gasification stage and in the water-gas shift reaction stage, obviating the need to generate or supply additional high pressure steam. Typically, in the cooling stage, condensed water is removed from the synthesis gas.

According to another aspect of the invention, there is provided a process for producing a synthesis gas derived product, which process includes

producing a synthesis gas in a process as hereinbefore described; and
in a synthesis gas conversion stage, converting the synthesis gas to a
synthesis gas derived product.

5 In one embodiment of the invention, the synthesis gas conversion stage is a Fischer-Tropsch hydrocarbon synthesis stage. However, it is to be appreciated that the synthesis gas conversion stage may be any synthesis stage requiring synthesis gas, such as a methanol, higher alcohol or oxoalcohol synthesis stage.

The Fischer-Tropsch hydrocarbon synthesis stage may be provided with any suitable reactor such as a tubular fixed bed reactor, a slurry bed reactor or an ebullating bed reactor. The pressure in the reactor may be between 1 bar and 100 bar, while the temperature may be between 200 °C and 380 °C. The reactor will thus contain a Fischer-Tropsch catalyst, which will be in particulate form. The
15 catalyst may contain, as its active catalyst component, Co, Fe, Ni, Ru, Re and/or Rh. The catalyst may be promoted with one or more promoters selected from an alkali metal, V, Cr, Pt, Pd, La, Re, Rh, Ru, Th, Mn, Cu, Mg, K, Na, Ca, Ba, Zn and Zr. The catalyst may be a supported catalyst, in which case the active catalyst component, e.g. Co, is supported on a suitable support such as Al₂O₃, TiO₂, SiO₂,
20 ZnO or a combination of these.

The process may include removing sulphur compounds which may be present in the synthesis gas, prior to converting the synthesis gas to a synthesis gas derived product. Removing the sulphur compounds may include employing a
25 so-called Rectisol process, in which methanol is used to wash the synthesis gas.

Typically, in the Fischer-Tropsch hydrocarbon synthesis stage, a product gas is formed which includes methane. The methane may be removed from the product gas. In one embodiment of the invention, the methane is separated from
30 one or more condensate streams in a cold separation stage. The separated methane may be flared. Instead, the separated methane may be recycled to the

partial oxidation stage. The process of the invention may thus be characterised in that a reforming stage, e.g. a steam reforming stage, is not required downstream of the synthesis gas conversion stage, as the bulk of the methane formed during gasification of the carbonaceous material, and optionally the bulk of the recycled methane, is oxidised in the partial oxidation stage.

The invention will now be described, by way of example, with reference to the single accompanying diagrammatic drawing which shows a process in accordance with the invention for producing a synthesis gas derived product, the process including a process for producing synthesis gas.

Referring to the single drawing, reference numeral 10 generally indicates a process in accordance with the invention for producing a synthesis gas derived product.

The process 10, as shown by the simplified flow sheet in the drawing, includes a gasification stage 12, a partial oxidation stage 14, a water-gas shift reaction stage 16, a cooling stage 18, a synthesis gas sweetening stage 20, a synthesis gas conversion stage 22, a CO₂ removal stage 24 and a cold separation stage 26.

A coal feedstock line 28 and a steam and oxygen feed line 30 leads into the gasification stage 12, with an ash removal line 32 being provided to remove ash from the gasification stage 12. A raw gas line 34 leads from the gasification stage 12 to the partial oxidation stage 14. An oxygen feed line 36 also leads into the partial oxidation stage 14.

The partial oxidation stage 14 is connected to the water-gas shift reaction stage 16 by means of an upgraded synthesis gas line 38. A steam feed line 40 also leads into the water-gas shift reaction stage 16.

A hydrogen enriched synthesis gas line 42 connects the water-gas shift reaction stage 16 to the cooling stage 18. From the cooling stage 18, a high pressure steam line 44 and a condensate removal line 46 lead. In addition, the cooling stage 18 is connected to the synthesis gas sweetening stage 20 by means of a cooled synthesis gas line 48.

The synthesis gas sweetening stage 20 is provided with a naphtha removal line 50 and a sulphur removal line 52. A methanol feed line 54 leads into the synthesis gas sweetening stage 20.

The synthesis gas sweetening stage 20 is connected to the synthesis gas conversion stage 22 by a final synthesis gas line 56. An intermediate product line 58 and a reaction water removal line 60 lead from the synthesis gas conversion stage 22.

The CO₂ removal stage 24 is connected to the synthesis gas conversion stage 22 by means of a light gas line 62 and to the cold separation stage 26 by means of a CO₂ free light gas line 64. From the CO₂ removal stage 24, a CO₂ line 66 is taken, and from the cold separation stage 26, a condensate line 68 is taken. In addition, a methane-rich line 70 leads from the cold separation stage 26.

Typically, the gasification stage 12 comprises a plurality of moving bed gasifiers, such as Lurgi (trade name) moving bed gasifiers (not shown). These gasifiers produce synthesis gas with a CO : H₂ ratio between 1 : 1.7 and 1 : 2, which makes them ideal for the production of synthesis gas which is to be used in a Fischer-Tropsch hydrocarbon synthesis stage. Coarse coal is fed to the gasifiers as indicated by the coal feedstock line 28, with steam and oxygen being fed along the steam and oxygen feed line 30. Oxygen is required to combust some of the coal to supply energy for the endothermic gasification reactions. Typically, part of the steam that is used is generated in a gasifier jacket from boiler feed water that is fed to the jacket (not shown). The steam is at a pressure of 40 bar(gauge) and a

temperature of 390 °C, with the boiler feed water being at a pressure of 40 bar(gauge) and a temperature of 105 °C and the oxygen being at a pressure of 29 bar(gauge) and a temperature of 140 °C.

5 In each gasifier, within a gasifier bed, different reaction zones are distinguishable from top to bottom, namely a drying zone where moisture is released, a devolatilization zone where pyrolysis takes place, a reduction zone where mainly the endothermic reactions occur, an exothermic oxidation or combustion zone, and an ash bed at the bottom of the gasifier bed. As a result of the counter current mode of operation, hot ash exchanges heat with cold incoming reagents, such as steam and oxygen or air, while at the same time hot raw gas exchanges heat with cold incoming coal. This results in the ash and raw gas, respectively leaving the gasification stage 12 by means of the ash removal line 32 and the raw gas line 34, at relatively low temperatures compared to other types of gasifiers, which improves the thermal efficiency and lowers the steam consumption of the gasifiers.

20 In the pyrolysis zones of the gasifiers, tars, oils and pitches and the like are released. These pyrolysis products are not destroyed, in view of the lower operating temperatures of the Lurgi (trade name) gasifiers. The pyrolysis products can be used to create valuable co-products such as ammonia, sulphur, cresols and phenol.

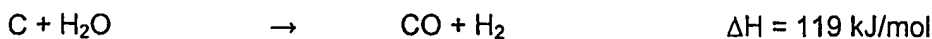
The following reactions take place in the gasifier:

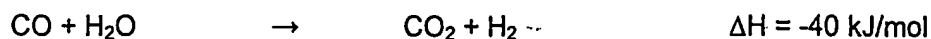
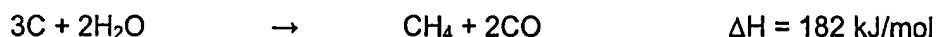
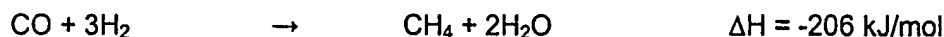
25

Combustion:



Reduction:



Water-gas shift:*Methane formation:*

10 The temperature profile in each gasifier varies as the coal moves through the different zones in the gasifier. In the gasification zone the temperature varies between 800 °C and 1200 °C. The raw synthesis gas leaving the gasification stage 12 is typically at a temperature of between about 460 °C and 500 °C, but may be lower.

15 The maximum temperature in the gasifiers is limited by the ash fusion temperature of the coal feedstock. Ash fusion creates removal problems of the ash at the bottom of the gasifiers. Owing to this limitation, the temperatures cannot be raised, causing more methane to form part of the raw synthesis gas than would be the case with higher temperatures.

20 In addition to H₂, CO, CO₂, H₂O and CH₄, the raw synthesis gas fed from the gasification stage 12 to the partial oxidation stage 14 also includes solid particles and tars and contaminants such as H₂S.

25 In the partial oxidation stage 14, the methane is thermally partially oxidised at a temperature of about 1300 °C in a combustion process with the oxygen fed to the partial oxidation stage 14 by means of the oxygen feed line 36. During the partial oxidation reaction, the methane is converted to H₂ and CO, thus providing an upgraded synthesis gas which includes less methane and more H₂ and
30 CO than the raw synthesis gas fed to the partial oxidation stage 14. The tar and solids present in the raw synthesis gas are also cracked and combusted rendering

the upgraded synthesis gas from the partial oxidation stage 14 free of these troublesome components.

For the process 10 of the invention, it may be necessary to employ a deposition stage (not shown) where soot particles are deposited on a collecting surface, to separate soot possibly formed in the partial oxidation stage from the upgraded synthesis gas before the upgraded synthesis gas enters the synthesis gas conversion stage.

In the water-gas shift reaction stage 16, the upgraded synthesis gas is mixed with high pressure steam at a temperature of about 600 °C to 850 °C, thereby subjecting the upgraded synthesis gas to a water-gas shift reaction to provide a hydrogen-enriched synthesis gas with a more desirable ratio of CO and H₂ of about 1 : 1.7. Thus, the hydrogen-enriched synthesis gas from the water-gas shift reaction stage 16 comprises more hydrogen and less CO than the upgraded synthesis gas fed to the water-gas shift reaction stage 16.

In the cooling stage 18, the hydrogen-enriched synthesis gas is cooled, with high pressure steam at a pressure of about 40 bar(gauge) being generated. During the cooling/steam generation process, water is condensed from the hydrogen-enriched synthesis gas, with the condensate being removed by means of the condensate removal line 46.

The cooled synthesis gas is transferred by means of the cooled synthesis gas line 48 to the synthesis gas sweetening stage 20. The synthesis gas sweetening stage 20 is a so-called Rectisol process intended to remove sulphur compounds, CO₂, higher hydrocarbons (naphtha) and HCN from the cooled synthesis gas.

In the synthesis gas sweetening stage 20, the cooled synthesis gas is further cooled in a series of heat exchangers (not shown) and any entrained gas

liquor is removed. Methanol is added to the synthesis gas to prevent icing as the cooled synthesis gas is still saturated with water. The cold synthesis gas is washed with more methanol in a pre-washed section of an absorber (not shown) which includes a pre-wash, a main wash and a fine wash section. In the pre-wash section, naphtha, HCN and water are removed to the bottom of the absorber while the synthesis gas moves up the absorber through the main wash and the fine wash sections. In the main wash and fine wash sections, methanol is used to remove sulphur gases and CO₂ from the synthesis gas. The final, cleaned synthesis gas leads from the top of the absorber and is fed to the synthesis gas conversion stage 22 along the final synthesis gas line 56. The sulphur compounds are removed from the synthesis gas sweetening stage 20 by means of the sulphur removal line 52, whilst the naphtha is passed to a refinery (not shown) along the naphtha removal line 50.

The synthesis gas conversion stage 22 makes use of high temperature Fischer-Tropsch conversion. The final synthesis gas is fed along the final synthesis gas line 56 to Fischer-Tropsch reactors where the hydrogen and carbon monoxide in the final synthesis gas react under pressure in the presence of a fluidised, iron-based catalyst at a modest temperature in the region of about 350 °C to yield a broad spectrum of hydrocarbons in the C₁-C₅₀ range. High value chemical components are produced simultaneously with synthetic oil. Oxygenated hydrocarbons and reaction water are also produced in the reactors:

The final synthesis gas is fed to the bottoms of the Fischer-Tropsch reactors at a rate which fluidises an iron catalyst bed allowing the Fischer-Tropsch reaction to proceed at the temperature of approximately 350 °C and a pressure of 20 bar. As the reaction is exothermic, steam-producing cooling coils (not shown) are present in the Fischer-Tropsch reactors to remove the heat of reaction. Fresh iron catalyst is added when required without disrupting the process to keep the conversion of the synthesis gas high and to ensure that the particle size distribution of the catalyst particles is kept constant, in order to avoid affecting the voidage of

the bed and the pressure drop through the bed. Typically, for high temperature fluidised bed Fischer-Tropsch reactors, the spread of products is more or less as shown in the following Table:

TABLE 1

Component	Mass % of product streams
CH ₄	8
C ₂ -C ₄	25
Fuel Range (Petrol, diesel, etc.)	62
Oxygenates	5

A variety of intermediate product streams, such as a decanted oil stream and a stabilised light oil stream, are removed from the synthesis gas conversion stage 22 for further working up in the refinery. The streams are generically indicated by the intermediate product line 58. Similarly, reaction water is removed from the synthesis gas conversion stage 22 by means of the reaction water removal line 60 for recovery of valuable products such as alcohols, ketones and organic acids.

A light gas product is withdrawn from the synthesis gas conversion stage 22 along the light gas line 62 and passed to the CO₂ removal stage 24. The CO₂ removal stage 24 is a so-called Benfield separation process which is employed to remove CO₂ from the light gas product, which typically comprises CO₂, H₂ and CH₄. As will be appreciated, CO₂ is formed in large quantities during Fischer-Tropsch synthesis as a result of the so-called water-gas shift reaction. Typically, the CO₂ makes up about 11 % of the light gas product withdrawn from the synthesis gas conversion stage 22.

In the CO₂ removal stage, the CO₂ gas is absorbed from the light gas in two trains using a solution of K₂CO₃ and diethanol amine. The light gas passes

through a K_2CO_3 wash column (not shown) followed by a diethanol amine wash column (not shown). The K_2CO_3 solution and the diethanol amine solution are regenerated in two separate regeneration columns (not shown). Typically, the CO_2 recovered during regeneration of the K_2CO_3 and diethanol amine solutions is released into the atmosphere, as indicated by the CO_2 line 66. Alternatively, the H_2/CO ratio of synthesis gas produced can be tailored by recycling CO_2 to the thermal partial oxidation stage.

A CO_2 free light gas stream from the CO_2 removal stage 24 is sent to the cold separation stage 26 by means of the CO_2 free light gas line 64. The purpose of the cold separation stage 26 is to separate out the methane, hydrogen, ethylene and propylene from the remaining gases in the CO_2 free light gas stream. This is effected by first cooling the CO_2 free light gas stream to a temperature of about $15\text{ }^\circ\text{C}$ at 27 bar and removing a C_3 hydrocarbon fraction and water. Further cooling to $-35\text{ }^\circ\text{C}$ and drying are effected in a second stage with a condensate stream containing a C_3 and C_2 fraction being removed for further purification in a de-ethaniser column (not shown). The remainder of the CO_2 free light gas stream is cooled again and the hydrogen and methane are separated. The hydrogen is sent to pressure swing absorbers (not shown) for purification. The methane, in the form of a methane condensate, is purified in a de-methaniser column (not shown) and then removed from the cold separation stage 26 as shown by the methane rich line 70. This methane can either be flared, or returned to the partial oxidation stage 14. The various condensate streams removed from the cold separation stage 26 are transferred to the refinery and are generically indicated by the condensate line 68.

The process 10 was simulated mathematically and the following Tables provide some of the salient information generated by the simulation. In the simulation, it was assumed that thirty-seven Lurgi (trade name) gasifiers are employed and that the steam that is generated during the process 10 is also used in the process 10. The Tables also include comparative information for a

conventional Lurgi (trade name) gasification process followed by quenching and cooling of raw synthesis with the conventional process being similar to the process 10 as far as the stages 20, 22, 24 and 26 are concerned. In the conventional process, however, methane from the synthesis conversion stage 22 is reformed in 5 autothermal reformers in the presence of steam and oxygen (at a carbon : steam ratio of 1 : 1.5) and recycled to join up with the raw synthesis gas from the gasifiers.

TABLE 2***Coal feedstock composition***

Component	Mass %
Fixed Carbon	46.03
Ash	23.67
Volatiles	22.61
Inherent moisture	4.18
Tar	3.50

TABLE 3***Synthesis gas composition and condition after various stages of the process 10***

	After gasification stage 12	After partial oxidation stage 14	After water-gas shift reaction stage 16
Temperature °C	404	1300	850.00
Pressure bar(g)	29.00	29.00	29.00
	Mole %		
H ₂ O	34.47	29.79	35.08
H ₂	20.74	31.89	30.71
CH ₄	10.64	0.22	0.46
CO	12.20	23.63	17.82
CO ₂	21.10	13.90	15.42
N ₂	0.60	0.57	0.51
TAR	0.26	0.00	0.00
Total (kmol/hr)	3898.00	4842.82	5424.84

TABLE 4***Final synthesis gas fed to synthesis gas conversion stage***

	Composition (mole %)	
	Conventional process	Process 10 of the invention
H ₂ O	0.17	0.20
H ₂	57.37	62.00
Argon	0.06	0.00
N ₂	1.96	1.03
CO	23.70	35.98
CO ₂	2.30	0.06
C ₁	14.34	0.93
Ethylene	0.01	0.01
C ₂	0.06	0.03
Propylene	0.02	0.04
Total (kmol/hr)	115442.80	99497.44

5

TABLE 5***Steam and oxygen requirements***

	Conventional process	Process 10 of the invention
HP steam (kgmol/hr)	87268.52	95941.00
O ₂ (kgmol/hr)	15693.54	20702.61

10

Although, according to the simulations, the conventional process produces 13.8 % more final synthesis gas feed to the synthesis gas conversion stage 22 for the same amount of gasifiers, there is 13 % more equipment capacity

available for the process 10 of the invention, as less methane is passed through the equipment. For the same equipment sizes downstream of the gasifiers, with the process 10 of the invention, as illustrated, more synthesis gas can be generated, using more gasifiers. Furthermore, for the process 10 of the invention, all the high pressure steam required during the process is generated by the process 10 whereas, for the conventional process, an additional 213 tons per hour coal must be combusted to generate sufficient steam for compression of oxygen and other purposes. The process 10 of the invention, as illustrated, also does not require a methane reformer although it is to be appreciated that, if additional methane, e.g. from natural gas is available, the operation of a methane reformer could be justified.

DATED THIS 28TH DAY OF OCTOBER 2002



ADAMS & ADAMS
APPLICANT'S PATENT ATTORNEYS

